Behaviour of Three Phase Fluidized Bed with Irregular Particles

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In partial fulfillment of the requirements

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By

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CERTIFICATE

This is to certify that the work in this thesis entitled "**Behaviour of Three Phase Fluidized Bed with Irregular Particles**" submitted by **Barsha Marandi (108CH024)** in partial fulfillment of the requirements of the prescribed curriculum for Bachelor of Technology in Chemical Engineering, Session 2008-2012, in the department of Chemical Engineering, National Institute of Technology, Rourkela, is an authentic work carried out by her under my supervision and guidance. To the best of my knowledge the matter embodied in the thesis is her bona fide work.

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ABSTRACT

The hydrodynamic characteristics- bed pressure drop, minimum fluidization velocity, bed expansion ratio and bed fluctuation ratio for a co-current three phase fluidized bed with irregular particles have been determined. Experiments were performed with air, water and dolomite as the gas, liquid and solid phases respectively using liquid as the continuous phase and gas as the discontinuous phase. Dolomite particles of 2.18 mm and 3.075 mm diameter were used as the bed material. Variation of bed pressure drop, minimum fluidization velocity, bed expansion ratio and bed fluctuation ratio with various parameters were investigated.

The pressure drop decreased with gas and liquid velocity, but it increased with static bed height and particle size. Minimum fluidization velocity was found to be independent of initial static bed heights. It decreased with liquid and gas velocity and increased with particle size. Bed expansion ratio increased with gas and liquid velocity, but it decreased with particle size. It remained the same for different static bed heights. Bed fluctuation ratio increased with increasing static bed heights, gas, and liquid velocity, but it decreased with particle size.

Keywords: Fluidization, three-phase fluidized bed, irregular particles, hydrodynamics, pressure drop, minimum liquid fluidization velocity, bed expansion ratio, bed fluctuation ratio.

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NOMENCLATURE

U_g	Superficial gas velocity, m/s
U_l	Superficial liquid velocity, m/s
U_{lmf}	Minimum liquid fluidization velocity, m/s
U_{gmf}	Minimum gas fluidization velocity, m/s
ΔP	Bed pressure drop, kPa
R	Bed expansion ratio
r	Bed fluctuation ratio
H_e	Expanded bed height, m
H_s	Static bed height, m
H_{max}	Maximum bed height, m
H_{min}	Minimum bed height, m
g	Acceleration due to gravity, m sec^{-2}
M_s	Mass of solids in the bed, kg
A	Area of the fluidizing column, m^2
D_p	Particle size, mm

Greek symbols

ρ	Density, kg m^{-3}
μ	Viscosity, Ns m^{-2}
ρ_s	Density of solid, kg m^{-3}
ρ_l	Density of liquid, kg m^{-3}
ρ_g	Density of gas, kg m^{-3}
ε_s	Solid holdup
ε_l	Liquid holdup
ε_g	Gas holdup
ε	Porosity

CHAPTER 1

INTRODUCTION

The process in which a bed of solid particles is converted from static solid-like state to dynamic fluid-like state when a fluid (liquid or gas or both) is passed up through the solid particles is termed as fluidization. If a fluid is passed through a bed of fine particles at a low flow rate, the fluid just percolates through the void spaces between the stationary solid particles. This condition is called the fixed bed. With increased flow rate, particles move away from each other and a few vibrate and move in restricted regions. This condition is called the expanded bed. At a still higher velocity, a point is reached where all the particles are just suspended by the upward flowing fluid. At this point, the frictional force between solids and fluid just counterbalances the weight of the particles, the vertical component of the compressive force between adjacent particles disappears, and the pressure drop through any section of the bed about equals the weight of fluid and particles in that section. The bed is considered to be just fluidized and is referred to as minimum or incipient fluidization (Chidambaram, 2011). At this stage, the bed is said to be fluidized and will exhibit fluidic behaviour. Under fluidized state, a bed of solid particles will behave as a fluid, like a liquid or gas.

1.1 Gas-Liquid-Solid Fluidized Bed

A gas-liquid-solid fluidized bed is a batch of solid particles suspended in a column by insertion of liquid and gas. In recent years, gas-liquid-solid fluidizing beds have developed as one of the most promising means for three phase operations. They are regarded immensely important in chemical and bio-chemical industries, treatment of waste water and other biochemical processes. Due to its increasing importance, its hydrodynamic characteristics such as the pressure drop, minimum fluidization velocity, bed expansion and fluctuation have to be investigated in order to provide the basic information required for the design of such fluidized beds (Chidambaram, 2011).

1.2 Advantages of Three Phase Fluidized Beds

The three phase fluidized beds are progressively being used as reactors as it can overcome some inherent drawbacks of conventional reactors and add more advantages than the conventional ones. Some of the advantages of three phase fluidized bed reactor are as follows:

- The smooth, fluid like flow of particles allows smooth, continuous, automatically controlled operations with simple handling.
- Achieves high turbulence, better mixing flexibility, heat recovery, and temperature control.
- Prevents formation of local hotspots.
- It is suitable for large scale operations.
- Better gas phase distribution is obtained, thus, creating more gas-liquid interfacial area.
- Ability to continuously withdraw products and introduce new reactants into the reaction vessel allows production to be more efficient due to the removal of start-up conditions as in case of batch processes.
- Allows use of fine catalyst particles; minimizing the intra-particle diffusion. Smaller the particle larger is the surface area enabling more intimate contact of phases and enhancing the reactor performance.
- Can be effectively used for rapidly deactivating catalyst and three phase reactions where the catalyst and the reactant are both in solid phases. (e.g. catalytic coal liquefaction).
- Fluidized bed technology is more economically, operationally and environmentally beneficial over other traditional technologies (Levenspiel et al., 1991 and Pandey, 2010).

1.3 Applications of Three Phase Fluidized Beds

The three phase fluidized bed has emerged in recent years as one of the most promising devices for three phase operation. In the past few decades, it has found immense applications in various fields like pharmaceutical, chemical, petrochemical, biochemical processing, metallurgical, and mineral processing. Fluidized beds serve many purposes in the industry, such as facilitating catalytic and non-catalytic reactions, mass and heat transfers, catalytic cracking, pyrolysis, and combustion. Three-phase fluidized beds have been applied successfully to many industrial processes such as in Hydrogen-oil process for hydrogenation and hydro-desulphurization of residual oil, the H-coal process for coal liquefaction, and in Fischer-Tropsch process. Some more applications of fluidized bed are as follows:

- Turbulent contacting absorption for flue gas desulphurization.
- Bio-oxidation process for wastewater treatment.
- Physical operations as drying and other forms of mass transfer.
- Biotechnological processes such as fermentation and aerobic wastewater treatment.
- Methanol production and conversion of glucose to ethanol.
- Pharmaceuticals and mineral industries.
- Oxidation of naphthalene to phthalic anhydride (catalytic).
- Coking of petroleum residues (non-catalytic) (Pandey, 2010).

1.4 Drawbacks of Fluidized Beds

- It is difficult to describe flow of gas, with its large deviation from plug flow and bypassing of solids by bubbles results in an inefficient contacting system.
- The rapid mixing of solids in the bed leads to non-uniform residence time of solids in the reactor.
- Attrition of solids is one of the main disadvantages. Because of attrition, the size of the solid particles gets reduced, and the possibility for entrapment of solid particles with the fluid is more (Chidambaram, 2011).

1.5 Modes of Operation of Gas-Liquid-Solid Fluidized Bed

Based on the differences in flow directions of gas and liquid and in contacting patterns between the particles and the surrounding gas and liquid, several types of operation for gas-liquid-solid fluidizations are possible. Gas-liquid-solid fluidization is classified mainly into four modes of operation.

- Co-current three phase fluidization with liquid as the continuous phase.
- Co-current three phase fluidization with gas as the continuous phase.
- Inverse three phase fluidization.
- Fluidization by a turbulent contact absorber (Jena et al., 2008).

1.6 Experimental survey

Analysis of literature on hydrodynamics in gas-liquid-solid fluidization shows that substantial work has been carried out, and the detailed research investigations were based on experiments carried out in small-scale columns. Extensive studies have been made into many aspects of three-phase fluidization and extensive reviews are available since 1970. However, three phase systems are complex and many questions are still unanswered. Although wide explorations on the hydrodynamics of three-phase fluidized beds have been made, a major problem that limits their industrial application implicates the difficulties in scaling-up the results from small laboratory units to larger industrially significant units. It is common for results, found with small-scale tests units, to be impracticable when the unit size is amplified. This problem is due to the materials which have been widely used in the previous work: air, water, and spherical glass-beads, while industrial applications mostly involve non-spherical particles at elevated temperatures and pressures. These temperatures and pressures result in differences in the properties of fluids, from that of air and water, at normal laboratory conditions (Jena, 2009 b).

For successful operation of gas-liquid-solid fluidized beds, it is necessary to have knowledge of minimum liquid fluidization velocity. The minimum liquid velocity required to achieve fluidization was determined by a plot of pressure drop across the bed vs. superficial liquid velocity at constant gas flow rate. It is the point at which the pressure drop does not change significantly. Thus, the velocity at which a break in curve occurred indicated minimum fluidization velocity. The minimum liquid fluidization velocity was also determined visually as either the velocity at which the bed first began to expand or as the velocity at which particles at the top of the bed started to exchange their position. Gas holdup has also been an important parameter where mass transfer is the rate limiting step. The following equations have typically been used to determine the volume fraction (holdup) of each phase in the three phase fluidized bed:

$$\boldsymbol{\varepsilon_s + \varepsilon_l + \varepsilon_g = 1} \qquad \ldots\ldots\ldots\ldots\ldots\ldots\ldots(1)$$

$$\boldsymbol{\Delta P = gH_e(\rho_s\varepsilon_s + \rho_l\varepsilon_l + \rho_g\varepsilon_g)} \qquad \ldots\ldots\ldots\ldots\ldots\ldots\ldots(2)$$

$$\boldsymbol{\varepsilon_s = M_s / \rho_s A H_e} \qquad \ldots\ldots\ldots\ldots\ldots\ldots\ldots(3)$$

The expanded bed height was obtained either visually or from the measured pressure drop gradient. Bed pressure drop was visually determined from the U-tube manometer using carbon tetrachloride or mercury as the manometric fluid which were connected to pressure ports at the top and at the bottom of the column. Gas holdup was also determined by isolating a representative portion of the test section by simultaneously shutting two quick closing valves and measuring the fraction of the isolated volume occupied by the gas. More direct and promising methods of measuring the local gas holdup which have been used were electro-resistivity, electro conductivity methods, γ - ray transmission measurements and radioactive tracer techniques.

In the above mentioned literature, the solid phase used were – spherical particles like glass beads, steel balls, plastic beads and other spherical catalyst particles; cylindrical particles like aluminium cylinders and PVC cylinders; other cylindrical catalyst particles and irregular particles like sand, irregular gravel, quartz particles whose sphericity ranged from 0.7 to 1.0 approximately (Jena et al., 2009 c).

The bed height was determined from static pressure profiles up the entire height of the column or visually. From the pressure profiles, bed height can be was taken as the point at which a change in the slope was observed (Jena, 2009 b). Bed expansion ratio is the ratio of the average height of a fluidized bed to initial static bed height at a particular flow rate of the fluidizing medium above the minimum fluidizing velocity. It is an important parameter for fixing the height of fluidized bed required for a particular service. It is often determined visually. The expansion ratio of a fluidized bed depends on excess gas velocity, particle size, and initial bed height. Bed expansion is substantially greater in a two-dimensional bed than in a three-dimensional one. Bed expansions reported by different investigators have different meanings because of varied methods of measurement adopted (Singh et al., 1999).

Bed fluctuation ratio has been broadly used to quantify fluidization quality. It is the ratio of the highest and lowest levels, which the top of the fluidized bed occupies for any particular gas flow rate above the minimum fluidization velocity. The fluctuation ratio is maximum at a particular velocity, for a particular bed and then it either decreases (due to slug formation) or remains constant at higher velocity ratio thereafter (Singh et al., 2006). The fluctuation is often determined visually.

1.7 Flow Regime

In order to attain a stable operation with several operating variables, an understanding of the flow regimes in a fluidized bed is important. The separation between the flow regimes is still not defined properly. Up to date, seven different flow regimes for gas-liquid-solid co-current fluidized beds have been found, which are (Jena, 2009 b):

- **Dispersed bubble flow** – It generally relates to low gas velocities and high liquid velocities. In this flow, small bubbles of uniform size occur. Bubble conjoining is low in spite of high bubble frequency.

- **Discrete bubble flow** – It is the condition in which low liquid and gas velocities occur. In such a flow, bubble size is small with lower bubble frequency.

- **Coalesced bubble flow** – It is the flow in which low liquid velocities and intermediate gas velocities operate. In this case, the bubble size is big with enhanced bubble coalescence.

- **Slug flow** – In this regime, bubbles occur in the shape of large bullets with a diameter of that of the column and length exceeding the column diameter. Some smaller bubbles can also be observed in the wakes of the slugs. This flow has limited applications.

- **Churn flow** – It resembles the slug flow regime. With an increase in the gas flow, an increase in downward liquid flow near the wall is observed. It is explained as the alteration between bubbling and slug flow on the basis of two-phase fluidized systems.

- **Bridging flow** – It is an intermediate regime between the churn flow and the annular flow. In this flow, solids and liquid form “bridges” in the reactor that gets constantly broken and re-formed.

- **Annular flow** – At elevated gas velocities, a continuous gas phase develops in the core of the column.

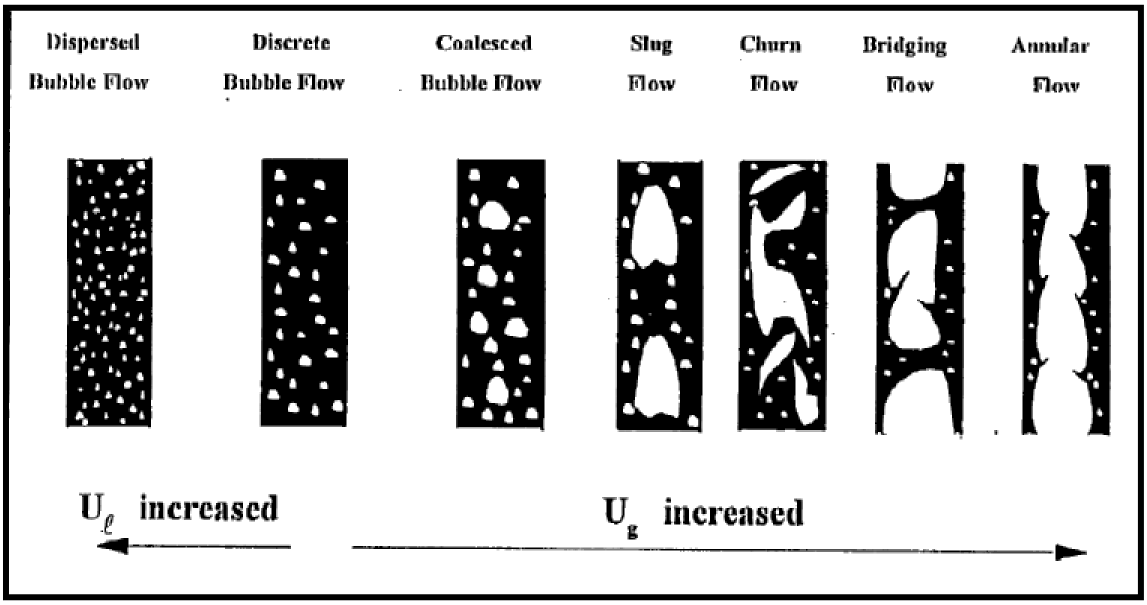

Fig.1. Flow regimes in gas-liquid-solid co-current fluidized bed

1.8 Variables Affecting The Quality of Fluidization

The quality of fluidization is influenced by the following variables (Chidambaram, 2011):

- **Fluid flow rate** – Flow rate should be enough to keep the solids in suspension, but it should not be high enough to result in fluid channelling.
- **Fluid inlet** – Inlet must be designed in such a way providing uniform distribution of the fluid entering the bed.
- **Particle size** – The quality of fluidization is greatly influenced by particle size. Particles of different sizes are grouped into Geldart's classification of particles. Particles having a wide range are easier to fluidize than the particles of uniform size.
- **Gas, liquid and solid densities** –Smooth fluidization is easily maintained when the relative density of the gas-liquid and the solid are closer.
- **Bed height** – With increasing bed height, it is difficult to maintain effective fluidization.

1.9 Some definitions in fluidization phenomena

1. Minimum Fluidization Velocity (U_{mf}) – The minimum superficial velocity at which the bed gets fluidized. At this velocity weight of the bed just gets counterbalanced by the pressure of the fluid.

2. Bed Pressure Drop (ΔP) – Measures the total weight of the bed in combination with the buoyancy and phase holdups.

3. Bed Expansion Ratio (R) – It the ratio of fluidized bed height to the initial static bed height.

$$R = H_e/H_s$$

4. Bed Fluctuation Ratio (r) – It is the ratio of maximum and minimum levels of the bed attained during fluidization.

$$r = H_{max}/H_{min}$$

4. Gas Holdup (ε_g) – It measures the volume fraction of gas. It is the ratio of volume of gas to the total volume of bed.

$$\varepsilon_g = \frac{volumeof\ gas}{total\ bed\ volume}$$

5. Liquid Holdup (ε_l) – It measures the volume fraction of liquid. It is the ratio of volume fraction of liquid to the total bed volume.

$$\varepsilon_l = \frac{volumeof\ liquid}{total\ bed\ volume}$$

6. Solid Holdup (ε_s) – It measures the volume fraction of solid. It is the ratio of volume fraction of solid to the total volume of bed.

$$\varepsilon_s = 1 - \varepsilon_l - \varepsilon_g$$

7. Porosity (ε) – It is the volume occupied by both liquid and gas (Jena, 2009 b).

$$\varepsilon = \varepsilon_l + \varepsilon_g = 1 - \varepsilon_s$$

1.10 Scope and objective of the present investigation

The aim of the present work has been précised below:

- Design and fabricate a fluidized bed system with an air sparger and distributor arrangement, which ensures less pressure drop across the column and uniform distribution of fluids.
- Hydrodynamic study of gas-liquid-solid fluidized bed with irregular particles- bed pressure drop, bed expansion, bed fluctuation, and minimum fluidization velocity.
- Variance of hydrodynamic properties with various parameters.

The present work is focussed on understanding the hydrodynamic behaviour in a three phase fluidized bed. Experiments were performed in a fluidized bed of height 1.4 m with a diameter of 0.1 m. Dolomites of particle size 2.18 mm and 3.075 mm are used as the solid phase. The fluidization operation has been carried out with co-current insertion of liquid (water) as the continuous phase and gas (air) as the discrete phase. Superficial velocity of both liquid and gas has been varied in the range of 0-0.148 m/s and 0-0.153 m/s. The static bed heights of the solid phase in the fluidized bed used are taken as 0.175 m, 0.216 m, 0.256 m and 0.296 m.

1.11 Thesis Layout

The second chapter provides a comprehensive literature survey related to the hydrodynamic characteristics. It includes the experimental as well as the computational aspect of gas-liquid-solid fluidization. The third chapter deals with the experimental setup and procedure used to study the hydrodynamic properties. Chapter four deals with the results attained from the observations and discussing the interpretation of the analysis. Chapter five infers the conclusion drawn from the results obtained from the study and it also presents the scope for the future works.

2. EXPERIMENTAL WORK

2.1. Experimental Setup

The fluidized bed assembly consists of three sections, viz., the test section, the gas-liquid distributor section, and the gas-liquid disengagement section. Fig. 6 shows the schematic representation of the experimental setup used in the three-phase fluidization study. Fig. 2 gives the photographic representation of the test section. The test section is the main component of the fluidized bed where fluidization takes place. It is a vertical cylindrical Plexiglas column of 0.1 m internal diameter and 1.4 m height consisting of two pieces of perspex columns. The gas-liquid distributor is located at the bottom of the test section and is designed in such a manner that uniformly distributed liquid and gas mixture enters the test section. The distributor section made of Perspex is fructo-conical of 0.31 m in height and has a divergence angle of 4.5^{o}. The liquid inlet of 0.0254 m in internal diameter is located centrally at the lower cross-sectional end. The higher cross-sectional end is fitted to the test section, with a perforated distributor plate made of G.I. sheet of 0.001 m thick, 0.12 m diameter. The holes were in triangular pitch. This has been done to have less pressure drop at the distributor plate and a uniform flow of the gas-liquid mixture into the test section. To avoid this unequal distribution at the entrance of the test section, the distributor plate has been designed so that relatively uniform flow can be achieved throughout the cross-section. Fig. 3 (b) and 5(b) represent the photographic view of the gas-liquid distributor section and the distributor plate. An antenna-type air sparger (Fig. 5(a)) of 0.09 m diameter with 50 number of 0.001 m holes has been fixed below the distributor plate with a few layers of plastic and glass beads in between for the generation of fine bubbles uniformly distributed along the column cross-section of the fluidizer. The gas-liquid disengagement section at the top of the fluidizer is a cylindrical section of 0.26 m internal diameter and 0.34 m height, assembled to the test section with 0.08 m of the test section inside it, which allows gas to escape and liquid to be circulated through the outlet of 0.0254 m internal diameter at the bottom of this section.

For the measurement of pressure drop in the bed, the pressure ports have been provided at the top and at the bottom of the test section and fitted to the manometers filled with carbon tetrachloride as the manometric fluid.

Oil free compressed air from a centrifugal compressor (3 phase, 1 Hp, 1440 rpm) used to supply the air at nearly constant pressure gradient as fluidizing gas. The air was injected into the column through the air sparger at a desired flow rate using calibrated rotameter. Water was pumped to the fluidizer at a desired flow rate using water rotameter. Centrifugal pump of capacity (Texmo, single phase, 1 HP, 2900 rpm, discharge capacity of 150 lpm); was used to deliver water to the fluidizer. Water rotameter used was of the range 20 to 200 lpm. Air rotameter was of the range 10 to 100 lpm.

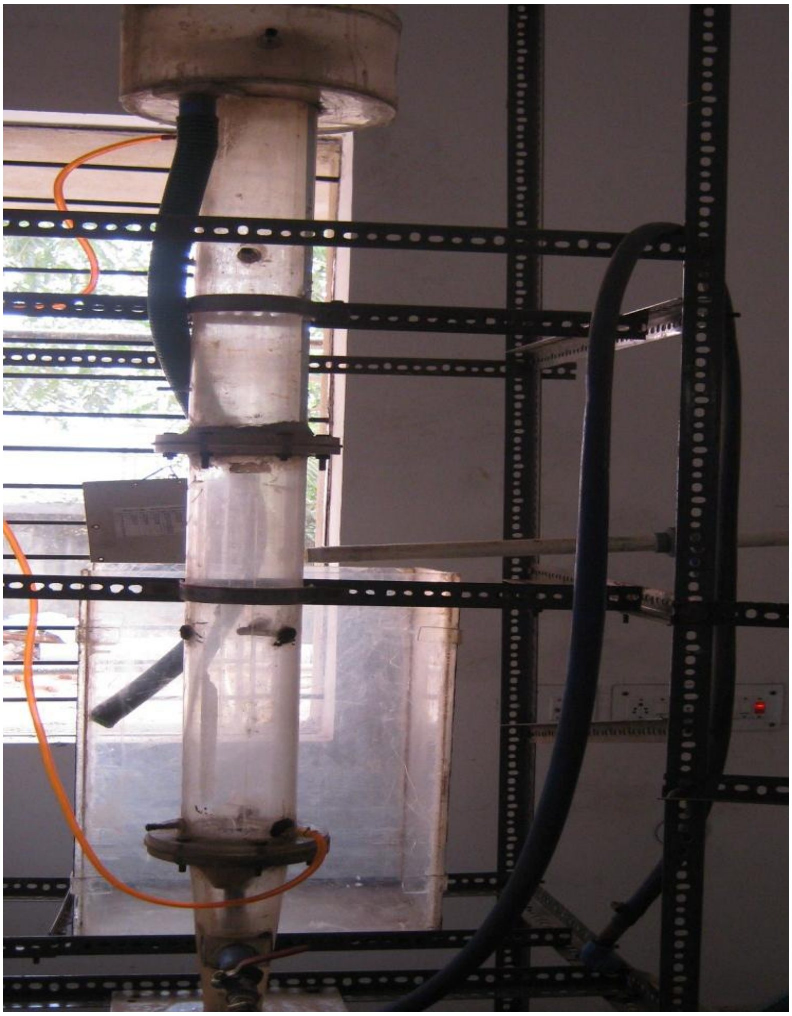

Fig.2. Photographic view of the test section

Fig.3. Photographic view of: (a) the disengagement section, (b) the gas-liquid distributor

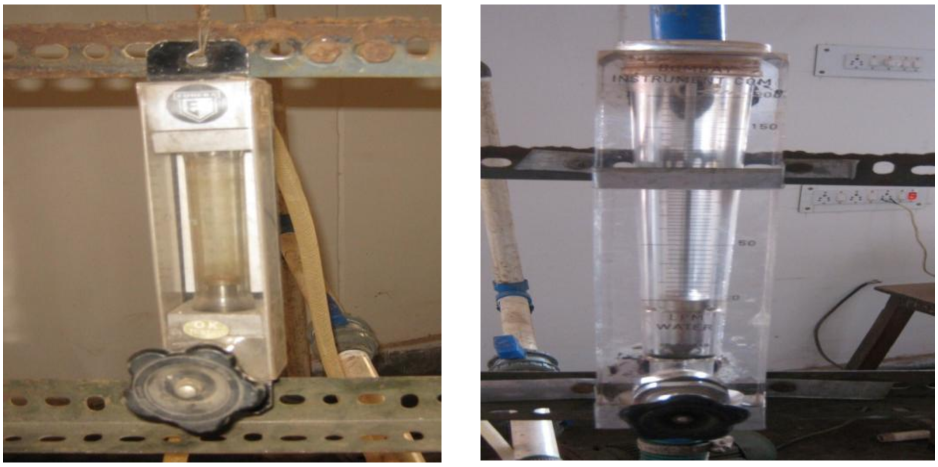

Fig.4. Photographic view of: (a) air rotameter, (b) water rotameter

Fig.5. Photographic view of: (a) air sparger, (b) distributor plate.

2.2 Experimental Procedure

The three phases present in the column were dolomite of particle size 2.18 mm and 3.075 mm, tap water and compressed air. The air-water flow were co-current and upwards. Material was first fed into the column and was adjusted for an initial static bed height. Water was pumped to the fluidizer at a desired flow rate. Then air was injected into the column through the air distributor. Approximately five minutes was allowed to attain steady state. Then, readings of each manometer were taken along with the bed expansion and fluctuation.

2.3 Scope of the Experiment

Table 1: Properties of Bed Material

Material	**d_p, mm**	**ρ_s, (kg.m^{-3})**
Dolomite	2.18	2940
Dolomite	3.075	2940

Table 2: Properties of Fluidizing Medium

Fluidizing Medium	**ρ (kg.m^{-3})**	**μ (Ns/m^2)**
Air at 25^{o}C	1.168	0.00187
Water at 25^{o}C	1.000	0.095

Table 3: Properties of Manometric Fluid

Manometric Fluid	**ρ (kg/m^3)**	**μ (Ns/m^2)**
Carbon tetrachloride	1600	0.09

Table 4: Operating Conditions

Superficial gas velocity	0-0.153 m/s
Superficial liquid velocity	0-0.148 m/s
Static bed heights	0.175-0.296 m

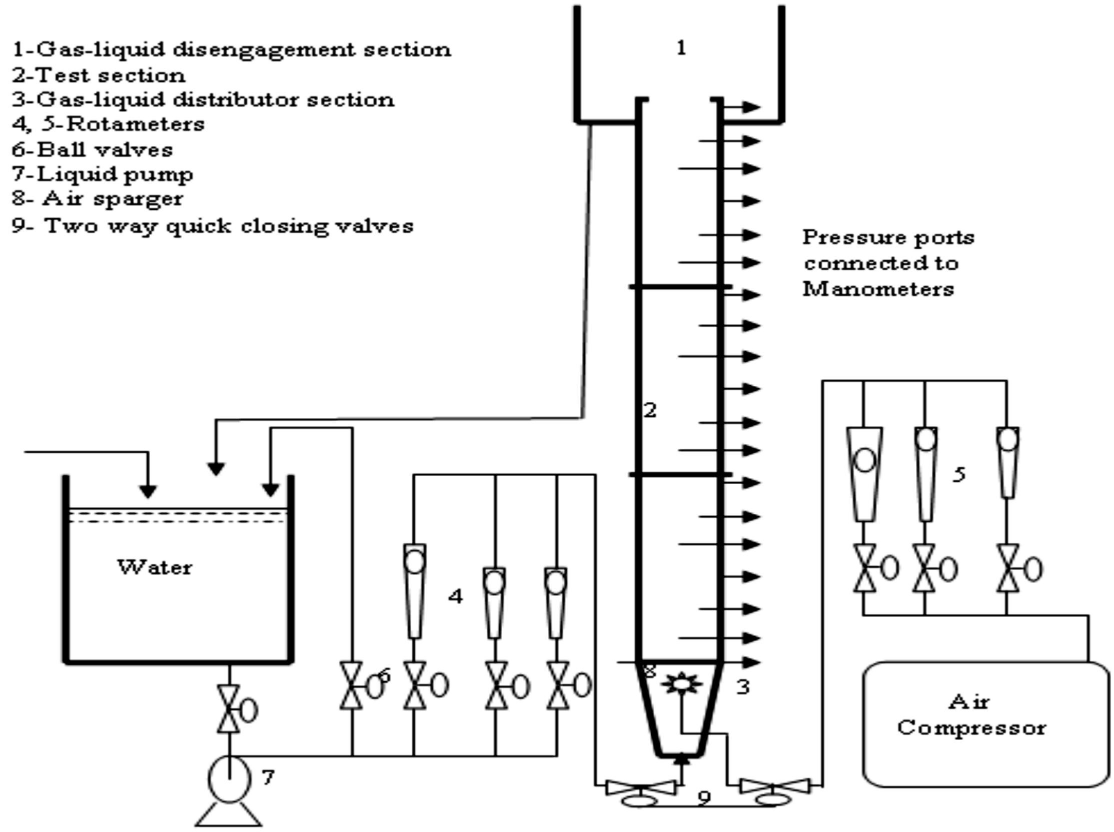

Fig. 6 Schematic diagram of the experimental setup

3. RESULTS AND DISCUSSION

In the present study, hydrodynamics of the gas-liquid-solid fluidized bed were studied. Dolomite, tap water and compressed air were used as the solid, liquid and the gas phases. The bed is likely to contain particles of varied shapes although the experimentally found sphericity of the total mass has been used to represent the particle shape. The bed expansion behaviour of such irregular particles is unlike that of the regular ones and higher expanded bed height is expected due to less sphericity and variation in shape of the particles. In the current chapter, an attempt has been made to acquire accurate knowledge of the hydrodynamic characteristics of fluidized bed by varying a large number of operating variables like liquid velocity, gas velocity, particle size and bed height. The hydrodynamic properties such as the bed pressure drop, minimum fluidization velocity, bed expansion ratio and bed fluctuation ratio have been studied and discussed in this chapter.

3.1 Pressure Drop

In this study, pressure drop is determined from manometer having carbon tetrachloride as manometric fluid in it, which is connected to ports at the top and at the bottom of the column. The column filled with solid particles up to a desired height and then occupied with water with the initial level of manometer adjusted to have zero. For gas-liquid-solid experiment with little flow of liquid close to zero, the air was introduced slowly and then increased gradually to the desired flow rate after which the liquid flow rate was increased, and the readings were noted down.

Fig. 7 shows the variation of pressure drop with superficial liquid velocity at different static bed heights with dolomite particles of 3.075 mm. As static bed height (or bed mass) increases, the pressure drop gradually increases for the fact that more drag force is needed to fluidize heavier beds so higher pressure drop is observed. For all the static bed heights, the minimum liquid fluidization velocity was found to be 0.042 m/s from this plot.

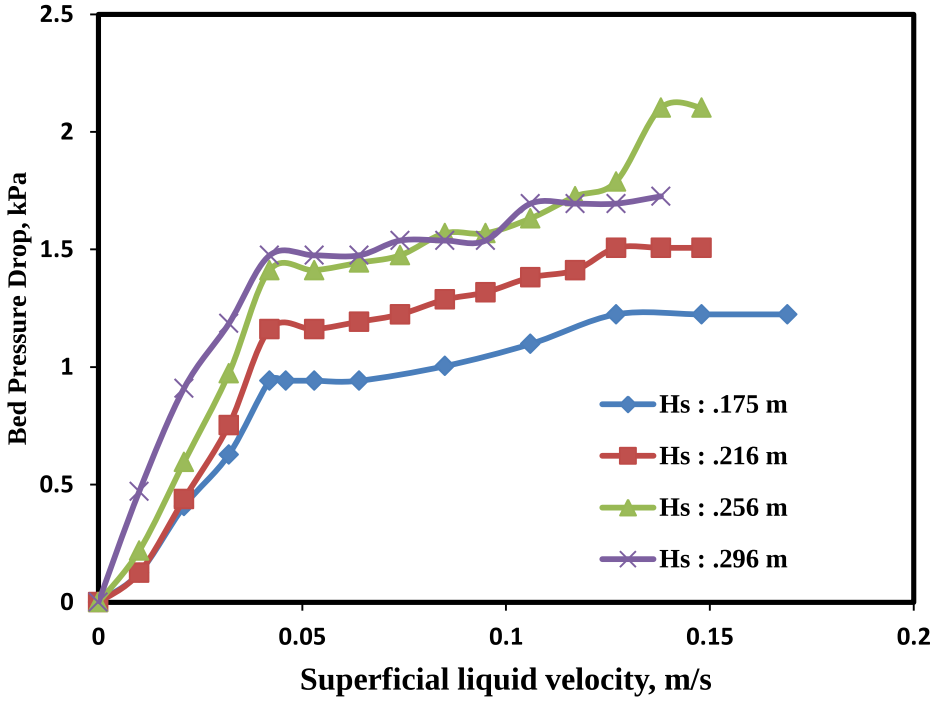

Fig.7. Variation of bed pressure drop with superficial liquid velocity at different static bed heights [U_g=0 m/s, D_p= 3.075 mm]

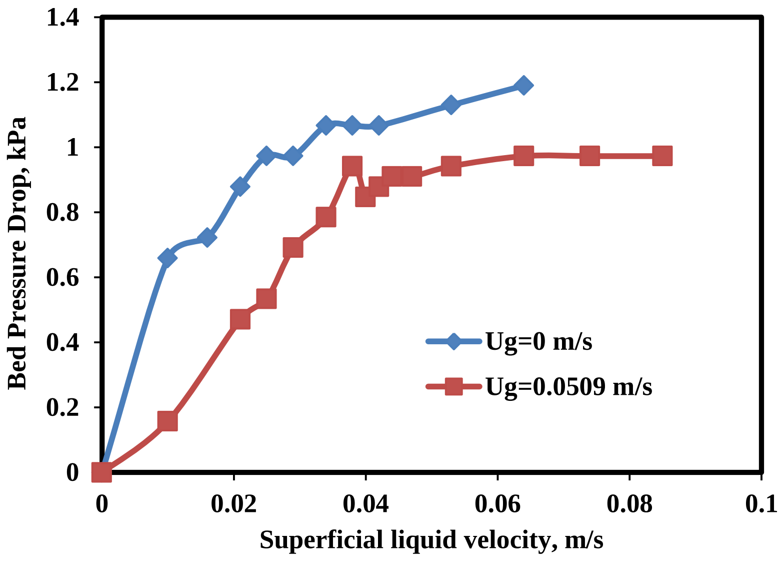

Fig.8. Variation of bed pressure drop with superficial liquid velocity at different superficial gas velocities [D_p= 2.18 mm, Hs=.175 m]

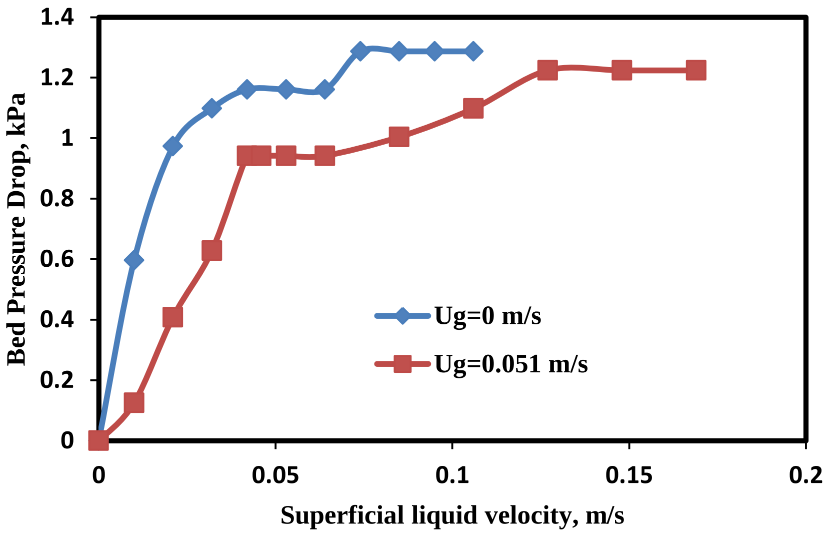

Fig.9. Variation of bed pressure drop with superficial liquid velocity at different superficial gas velocities [D_p= 3.075 mm, Hs=.175 m]

Fig. 8 and 9 exhibits the variance of bed pressure drop with superficial liquid velocity at particle size of 2.18 mm and 3.075 mm respectively. With increasing superficial gas velocities, it is observed that bed pressure drop decreases for both the particle sizes. This accounts for the point that with the introduction of gas, the force required to balance the weight of bed is reduced resulting in a decline in bed pressure drop. Moreover with higher particle size, the pressure drop is initially lower than the smaller particle size but gradually exceeds with increasing liquid velocity, interpreting the fact that more force is required to fluidize the compact bed formed due to small sized particles; but then as incipient fluidization is achieved, higher sized particle exert more force downwards owing to its weight. In addition to this, at higher gas velocity, the bed expansion is comparatively more, thus, pressure drop values are unavailable at higher liquid velocities as compared to liquid-solid fluidized bed.

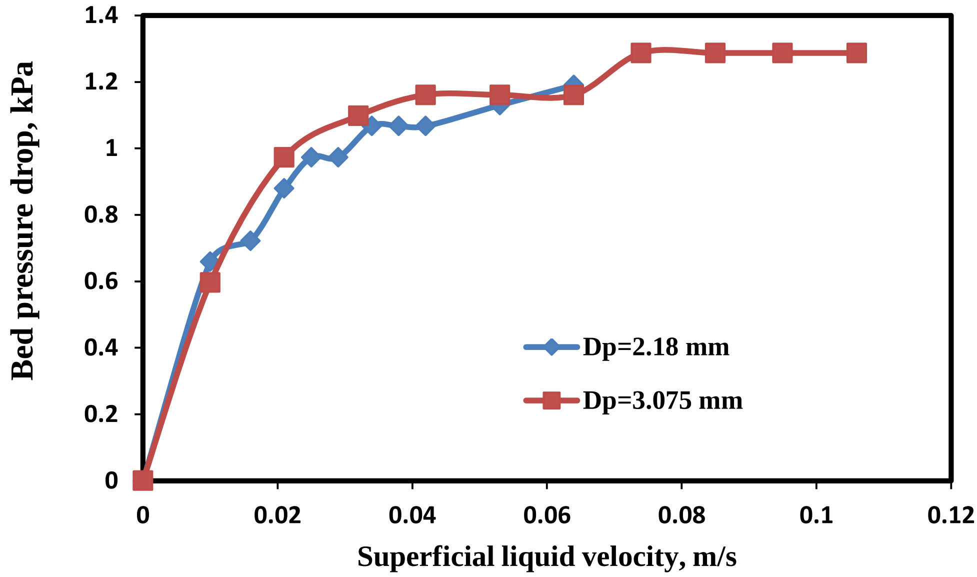

Fig.10. Variation of bed pressure drop with superficial liquid velocity for different particle sizes [U_g=0.051 m/s, Hs=.175 m]

Fig. 10 displays the effect of particle size on bed pressure drop. Initial pressure drop of lower sized particle is higher, but with a gradual increase of liquid velocity bed pressure drop for higher sized particles is higher as it exerts more downward force.

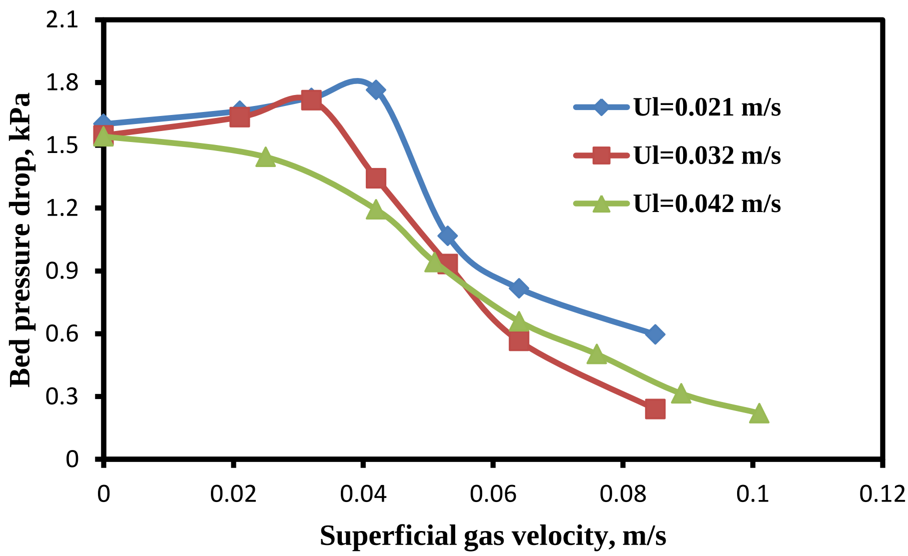

Fig.11. Variation of bed pressure drop with superficial gas velocity for different superficial liquid velocities [D_p= 3.075 mm, Hs=.216 m]

Fig. 11 shows the variation of bed pressure drop with superficial gas velocity for different superficial liquid velocities. For U_l=0.021 m/s and 0.032 m/s which are 50% and 75% of U_{lmf}=0.042 m/s for 3.075 mm, the bed pressure first increased until minimum fluidization was achieved and then decreased with further increase of gas velocity. For U_l=0.042 m/s, the pressure drop decreased with gas velocity. This accounts that with increase in gas holdup in the column, lesser drag force is required to suspend the bed of solids, thus lesser pressure drop is observed.

Although we use the manometer for measuring the bed pressure drop, but for a three phase system it does not represent the true frictional drag on the solid particles which holds the particles in suspended conditions as the hydrostatic pressure in the column changes due to gas holdup.

3.2 Minimum liquid fluidization velocity

Minimum liquid fluidization velocity has been determined visually in the present work. It is that velocity at which the bed first begins to expand or the velocity at which any particle within the bed continuously shifts position with neighbouring particles.

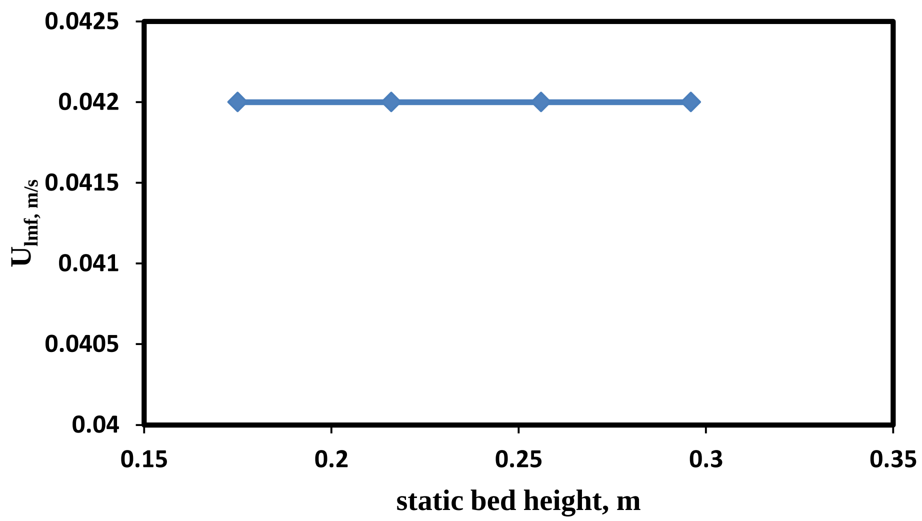

Fig.12. Variation of min. liquid fluidization velocity with initial static bed heights [D_p= 3.075mm, Ug=0 m/s]

In Fig. 12 the variation of minimum liquid fluidization velocity with initial static bed heights of .175m, .216 m, .256 m and .296 m has been shown at zero gas velocity. It is observed that there is no effect of initial static bed heights on minimum liquid fluidization velocity and a typical value of 0.042 m/s was observed for all the cases.

Table 5: Comparision of minimum gas fluidization velocity with superficial liquid velocity [Dp=3.075mm, Hs : .216 m]

U_l (m/s)	0.021	0.032	0.042
U_{gmf} (m/s)	0.04	0.028	0

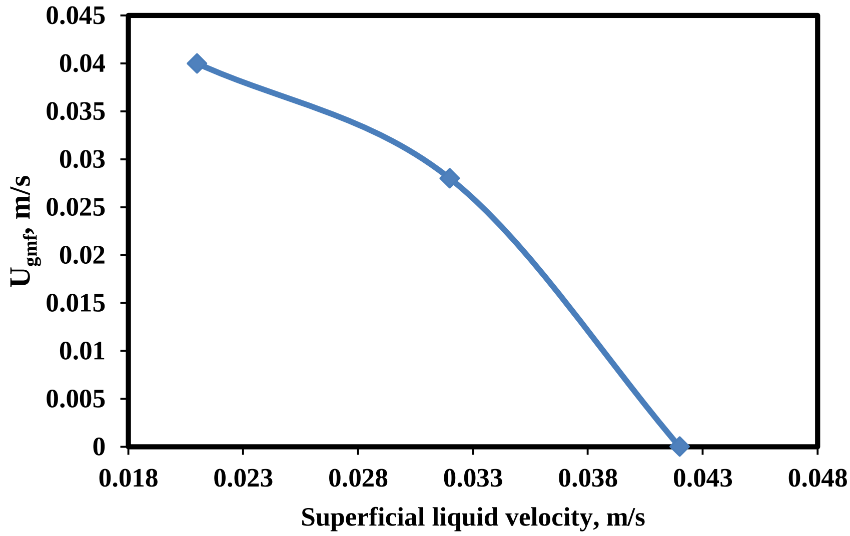

Fig.13. Variation of min. gas fluidization velocity with superficial liquid velocity [Dp=3.075mm, Hs : .216 m]

Fig 13 displays the variation of minimum gas fluidization velocity with superficial liquid velocities of 0.021 m/s, 0.032 m/s, 0.042 m/s for 3.075 mm particles and a static bed height of .216 m. The minimum gas fluidization velocities at these liquid velocities are given in table 5. It is spotted that with increasing superficial liquid velocity the minimum gas fluidization velocity decreases. This is due to the increased liquid holdup which makes it easier to achieve incipient fluidization.

Table 6: Comparision of min. gas fluidization velocity with superficial liquid velocity [Dp=2.18 mm, Hs : .175 m]

U_l (m/s)	0.021	0.032	0.042
U_{gmf} (m/s)	0.0212	0.017	0

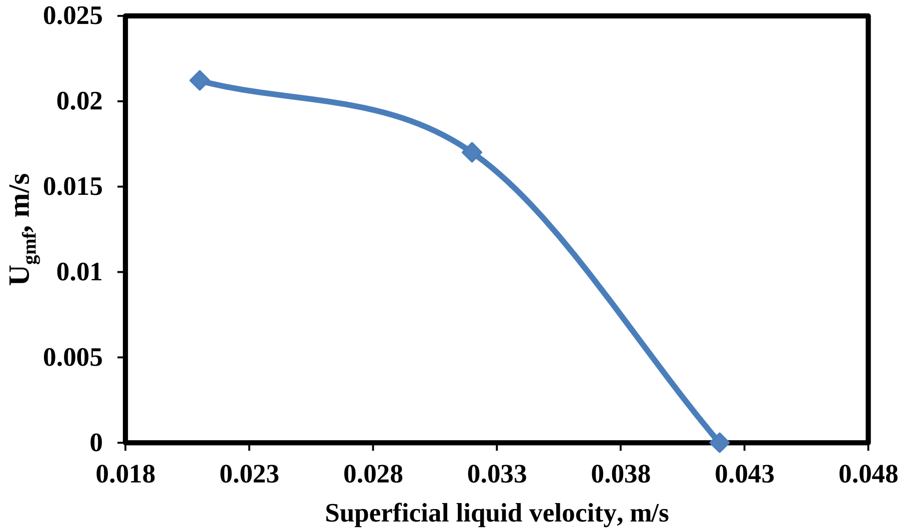

Fig.14. Variation of min. gas fluidization velocity with superficial liquid velocity [Dp=2.18 mm, Hs : .175 m]

Fig. 14 shows the influence of superficial gas velocity on minimum gas fluidization velocity. It is observed that the minimum gas fluidization velocity decreases linearly with superficial gas velocity. Values of minimum gas fluidization velocity obtained have been shown in table 6. This indicates that as the gas holdup in the column increases drag force needed to counterbalance the weight of the bed is reduced, thus easing incipient fluidization.

Table 7: Comparision of minimum liquid fluidization velocity with particle size[H_s=.175 m, U_g=0 m/s]

Particle size, mm	U_{lmf} (m/s)
2.18	0.04
3.075	0.042

Table 7 shows the effect of particle size on minimum liquid fluidization velocity. It is noticed that with larger particles the minimum liquid fluidization velocity is higher than the smaller particles. This indicates that smaller particles attain faster fluidization due to lesser drag force exerted by them.

3.3 Bed Expansion Ratio

In the present study, expanded bed height has been measured by visual observation. In visual observation, a very dilute bed which appears at the top of the three-phase region has been neglected and the height of the relatively dense bed has been reported as expanded bed height. The bed expansion study as carried out by varying liquid velocity (at a constant gas velocity) and particle sizes have been presented in terms of bed expansion ratio which is the ratio of the expanded bed height to the static bed height.

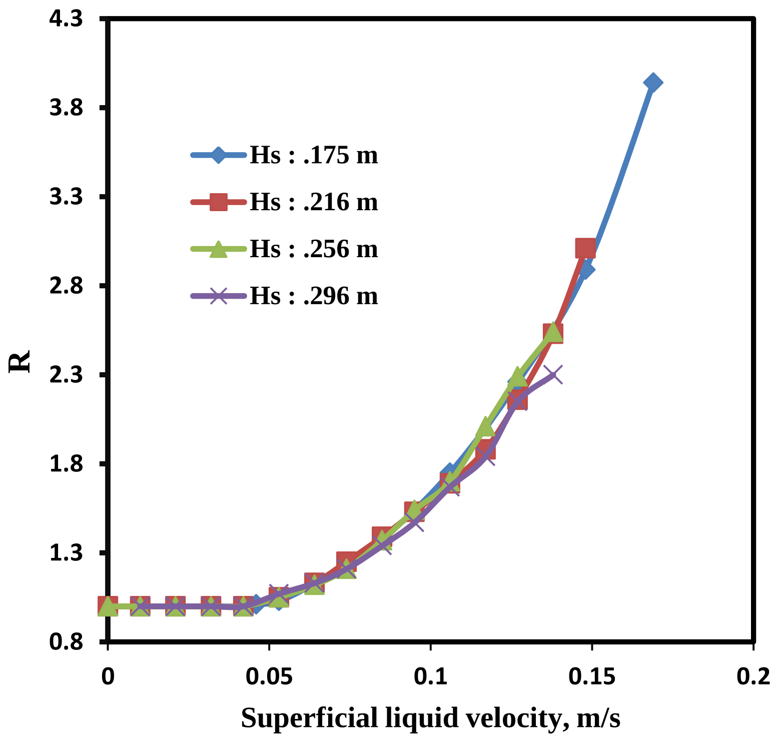

Fig. 15 Variation of bed expansion ratio with liquid velocity at different static bed heights [D_p=3.075 mm, Ug = 0 m/s].

Fig. 15 shows the variation of bed expansion ratio with liquid velocity at different static bed heights. It is observed that the bed expansion ratio is independent of static bed heights.

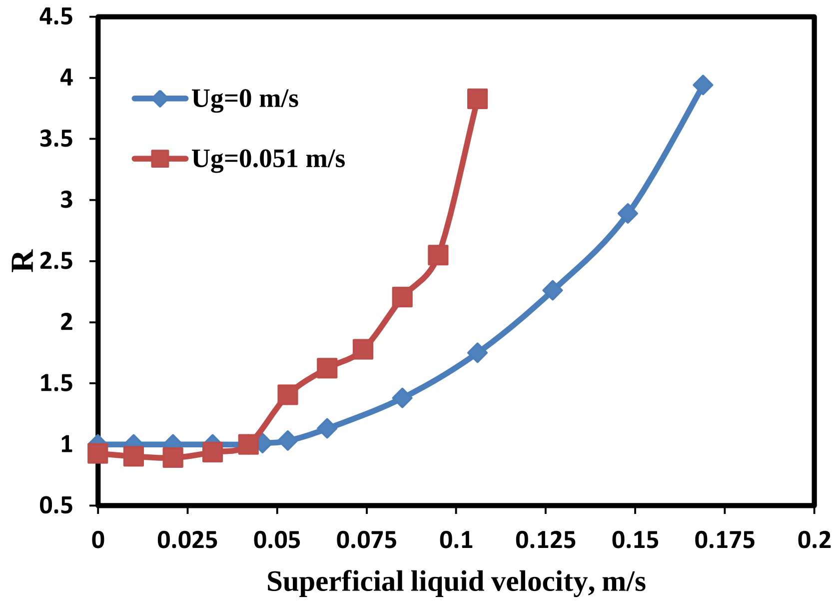

Fig. 16 Variation of bed expansion ratio with liquid velocity for different values of gas velocity at [Hs = 0.175 m, D_p = 3.075 mm].

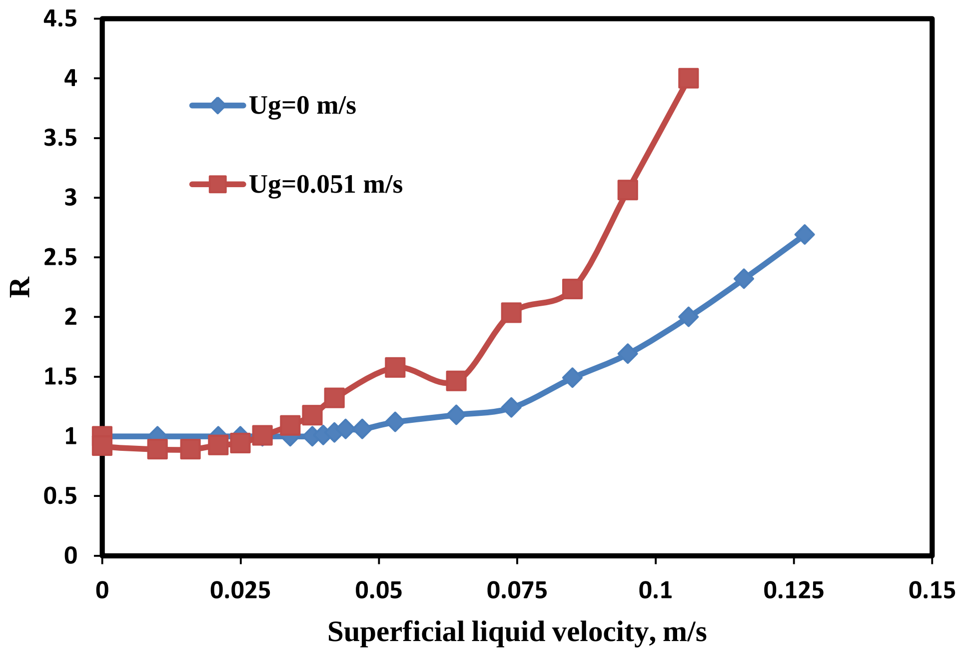

Fig.17. Variation of bed expansion ratio with liquid velocity for different values of gas velocity at [Hs = 0.175 m, D_p = 2.18 mm].

Fig. 16 and 17 shows the variation of bed expansion ratio with liquid velocity for different values of gas velocity of 3.075 mm and 2.18 mm particle sized dolomite respectively. Higher bed expansion ratio is achieved faster for solids operated at higher gas velocity. This is due to the increased gas holdup in the column which accounts for further expansion.

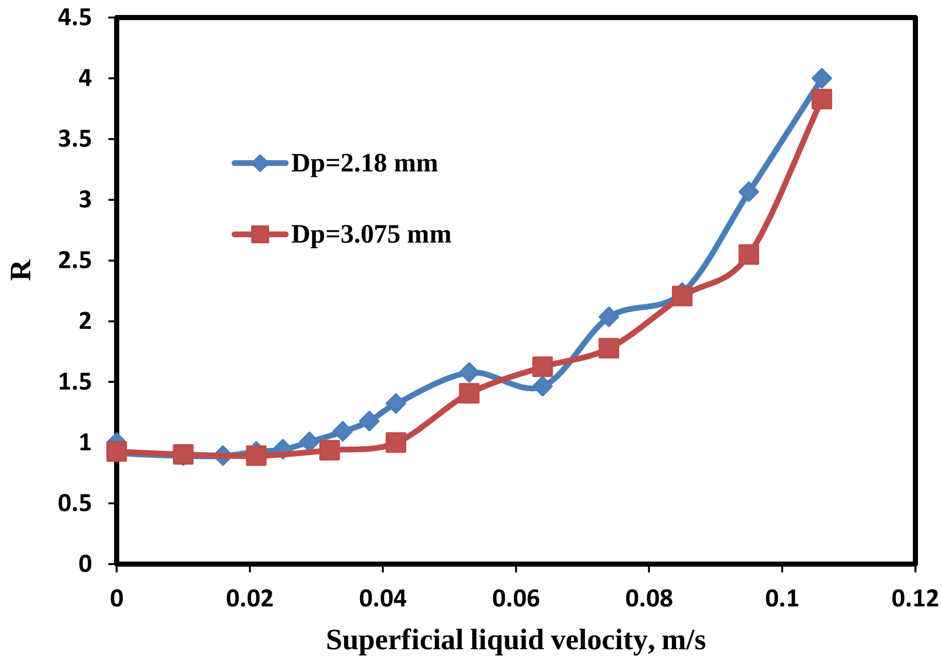

Fig.18. Variation of bed expansion ratio with liquid velocity for different particle sizes at [Hs = 0.175 m, U_g=0.051 m/s].

Fig. 18 shows the variation of bed expansion ratio with liquid velocity for different particle sizes. It is observed that for higher particle size bed expansion ratio is slightly lower than the smaller sized particle due to its greater weight.

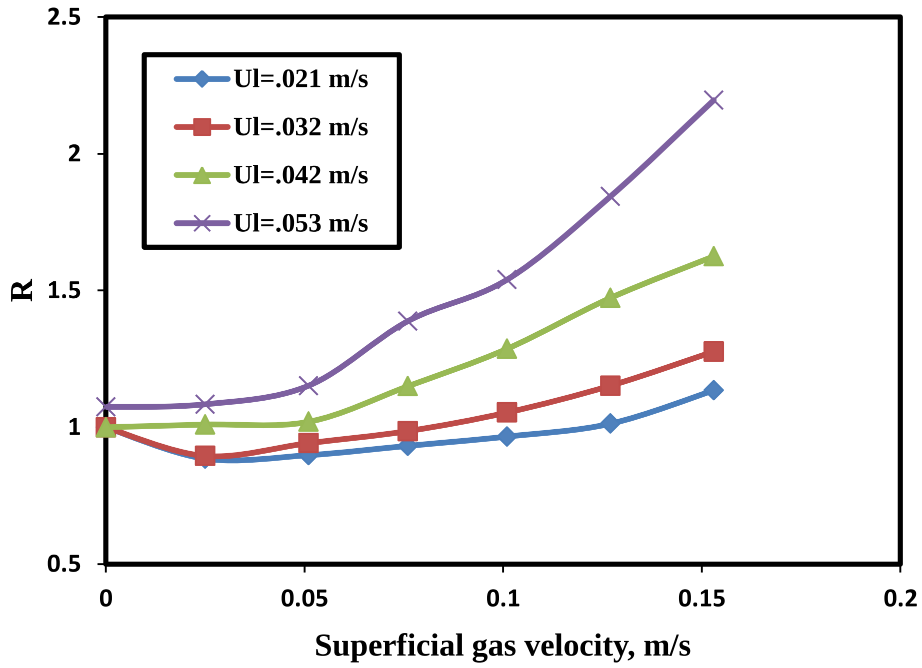

Fig.19. Variation of bed expansion ratio with gas velocity for different values of liquid velocity at [Hs =0.216 m, D_p = 3.075 mm].

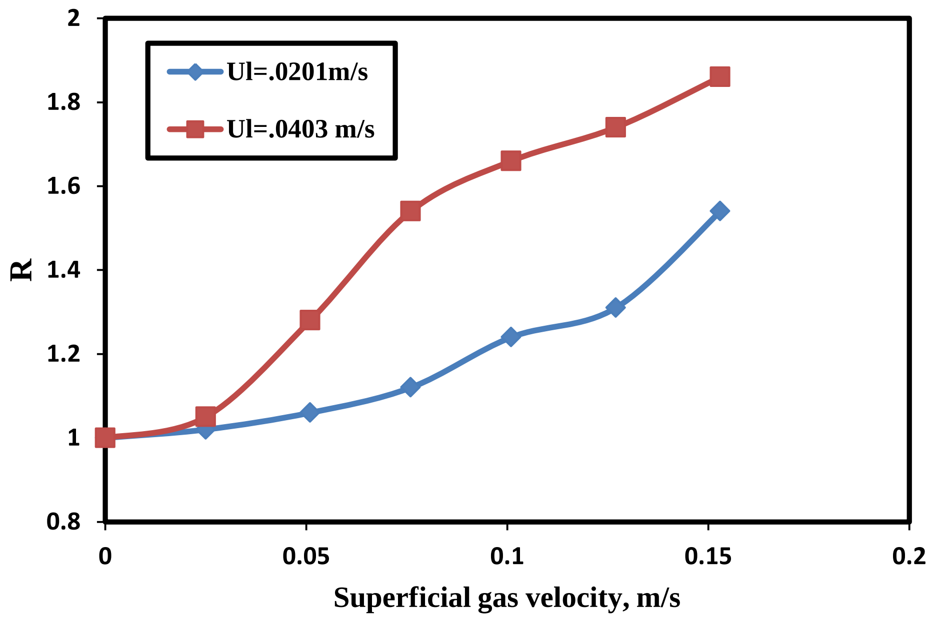

Fig.20. Variation of bed expansion ratio with gas velocity for different values of liquid velocity at [Hs = 0.175 m, D_p = 2.18 mm].

Fig. 19 and 20 shows the effect of liquid velocities on the bed expansion ratio for 3.075 mm and 2.18 mm paticle sizes. With increasing liquid velocities the bed expansion ratio increases due to the increased liquid holdup in the column leading to higher expansion.

3.4 Bed Fluctuation Ratio

In determining the bed fluctuation ratio, visual observation technique has been used in the current study. The bed fluctuation or the difference between the maximum and minimum levels of the bed height was noted down and the ratio between the two heights termed as fluctuation ratio was studied with various parameters. It is expected that the fluctuation ratio becomes maximum at a particular velocity, for a certain bed and then it either decreases (due to slug formation) or remains constant at subsequent velocities.

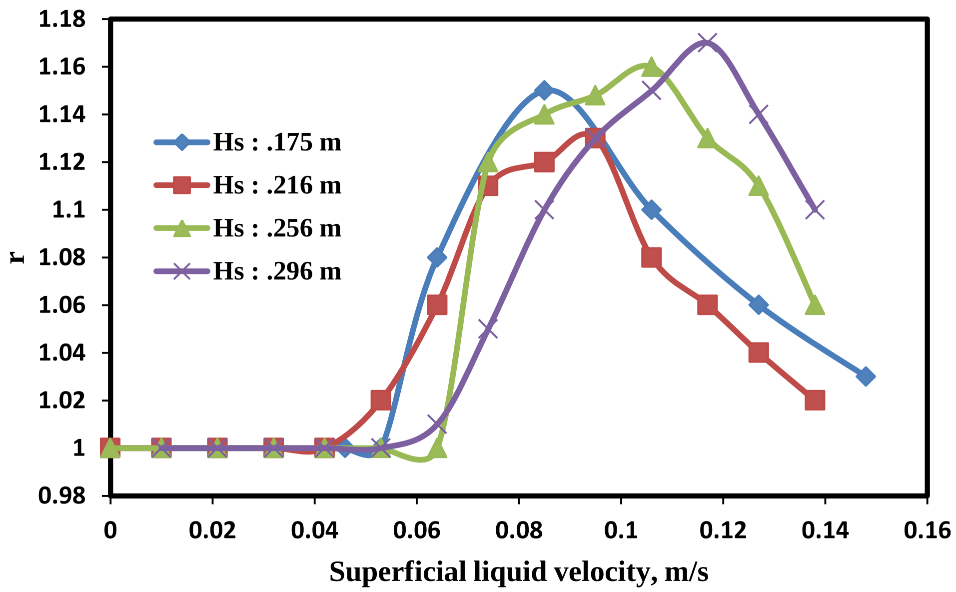

Fig.21. Variation of bed fluctuation ratio with liquid velocity at different static bed heights [D_p=3.075 mm, Ug = 0 m/s].

Fig. 21 shows the variation of bed fluctuation ratio with liquid velocity at different static bed heights. For all the bed heights a maximum fluctuation ratio is attained after which it is observed to decline. This indicates that initially the bed fluctuation increases with liquid velocity then it attains a maximum after which the fluctuation tend to decrease with higher liquid velocities as the entire bed then starts to expand. It is also observed that with increasing bed height the maximum fluctuation ratio attained in a particular system also increases.

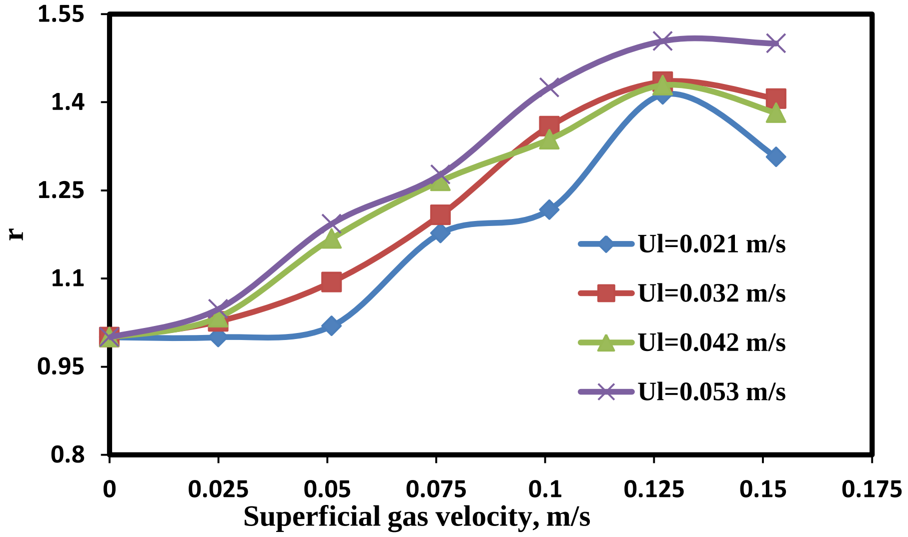

Fig.22. Variation of bed fluctuation ratio with gas velocity for different values of liquid velocity at [Hs = 0.216 m, D_p = 3.075 mm].

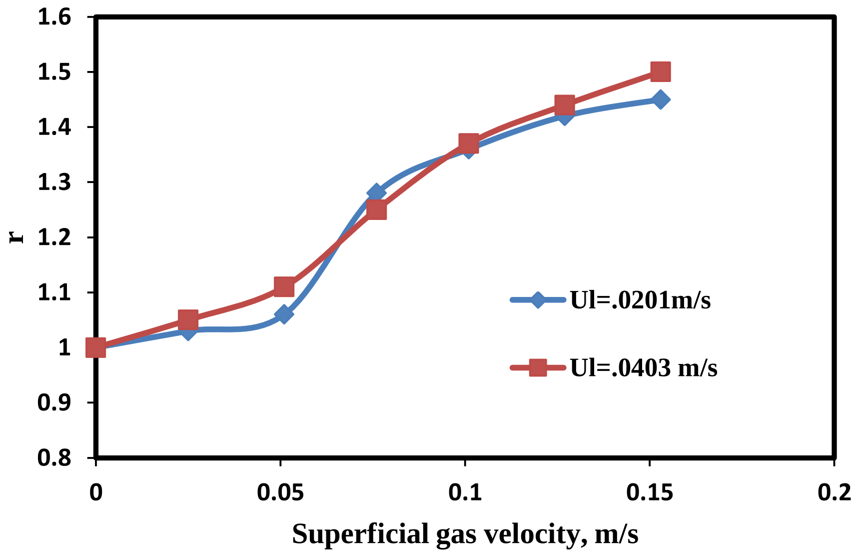

Fig.23. Variation of bed fluctuation ratio with gas velocity for different values of liquid velocity at [Hs = 0.175 m, D_p = 2.18 mm].

Fig. 22 and 23 depicts the variation of bed fluctuation ratio with gas velocity for different values of liquid velocity of 3.075mm and 2.18 mm particles respectively. For larger particles, the fluctuation ratio attains a maximum and then declines for lower liquid velocity. But as the liquid velocity is increased, the maximum then gradually advances to a constant. This indicates that with increased liquid holdup, beds containing larger particles tend to expand as a whole at higher liquid velocities. In contrast to this smaller particles attain higher fluctuation ratio faster than the larger particles and hence, bed levels at higher gas velocities cannot be measured as it fluctuates across the entire column at such high velocities.

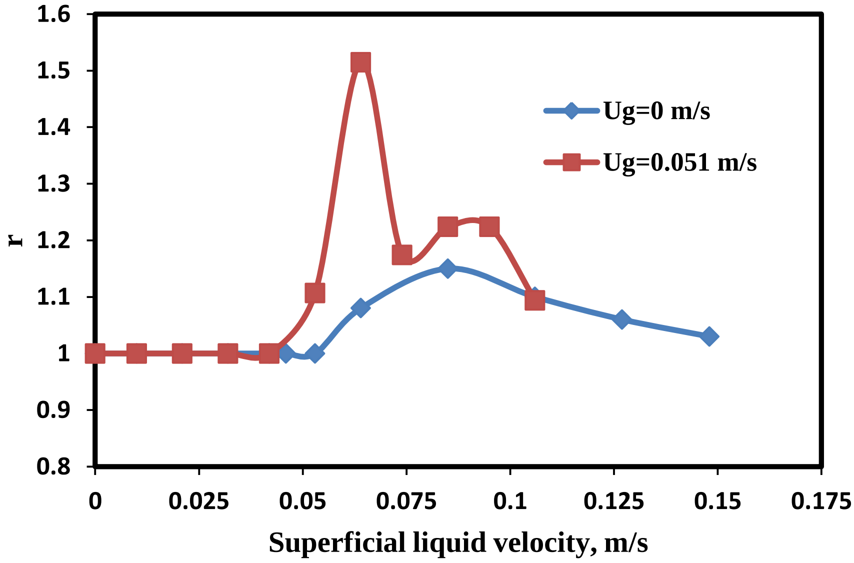

Fig. 24 Variation of bed fluctuation ratio with liquid velocity for different values of gas velocity at [Hs = 0.175 m, D_p = 3.075 mm].

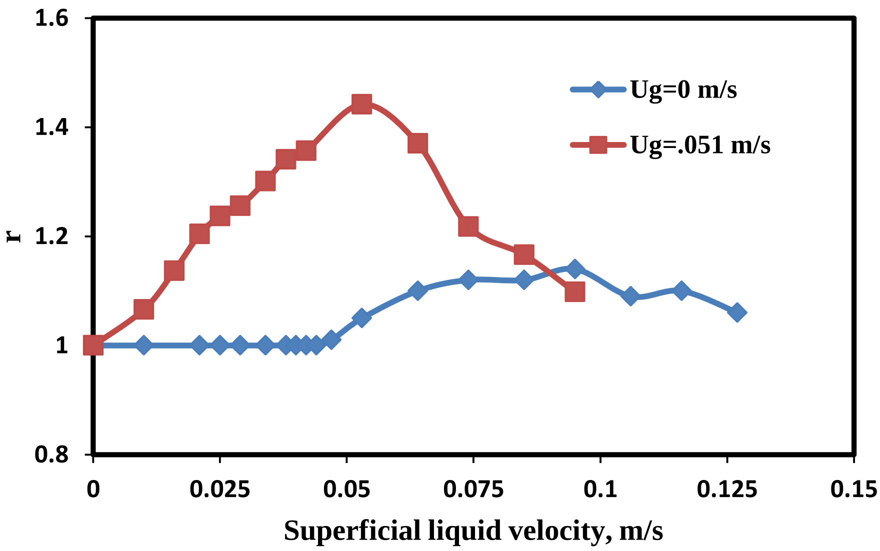

Fig. 25 Variation of bed fluctuation ratio with liquid velocity for different values of gas velocity at [Hs = 0.175 m, D_p = 2.18 mm].

Fig. 24 and 25 shows the variation of bed fluctuation ratio with liquid velocity for different values of gas velocity of 3.075 mm and 2.18 mm particles respectively. It is observed that the bed fluctuation ratio increases with increasing gas velocities. And for both the particle sizes, the fluctuation ratio attains a maximum and then decreases with increasing liquid velocities. This implicates that with increased gas holdup, the fluctuation in the bed increases.

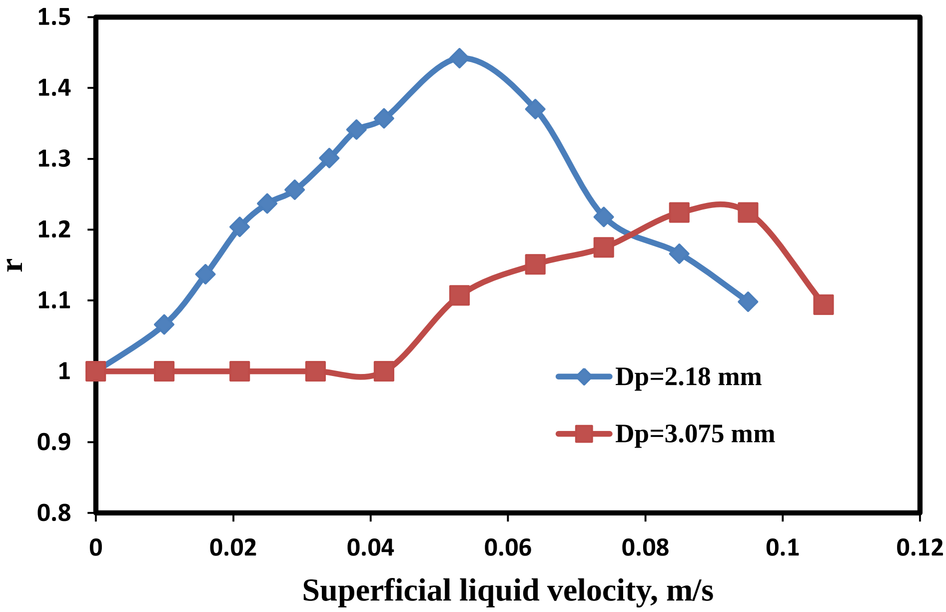

Fig. 26 Variation of bed fluctuation ratio with liquid velocity for different particle sizes at [Hs = 0.175 m, U_g=0.051 m/s].

Fig. 26 shows the variation of bed fluctuation ratio with liquid velocity for different particle sizes at constant gas velocity. It is observed that smaller particles fluctuate more as they reach a maximum fluctuation ratio earlier than the larger particles and the maximum fluctuation ratio achieved by the smaller particles is higher than that realized by larger particles. This indicates that bed fluctuation decreases with larger particles.

4. CONCLUSION

From the literature, it is revealed that fluidized beds are beneficial for efficacious gas-liquid-solid contacting process and can be used for waste water treatment, catalytic and non-catalytic reactors and in various chemical and bio-chemical processes. In the recent years, novel applications of fluidized bed systems are being discovered, which needs further understanding of the three phase fluidization systems. The problems related to the parameters affecting the hydrodynamics are still needed to be investigated. Even though a large number of experiments have studied the various hydrodynamic parameters of gas-liquid-solid fluidized beds, this complicated phenomenon has not yet been fully understood. Thus, many bleak areas still exist, which require further extensive fundamental studies in fluidized systems.

4.1 Pressure Drop

Effect of various parameters on bed pressure drop:-

- With increase in gas velocity, the pressure drop gradually decreases.
- As the static bed height increases, the pressure drop gradually increases.
- Bed pressure drop increases with particle size.
- With increasing liquid velocities, the pressure drop across the bed decreases.

4.2 Minimum liquid fluidization velocity

Effect of various parameters on minimum fluidization velocity:-

- Minimum liquid fluidization velocity is independent of initial static bed heights.
- With increasing liquid velocity, the minimum gas fluidization velocity decreases.
- Minimum liquid fluidization velocity increases with particle size.

4.3 Bed expansion ratio

Effect of various parameters on bed expansion ratio:-

- With increase in gas and liquid velocities, the expansion ratio increases.
- Bed expansion ratio is independent of initial static bed height.
- Bed expansion ratio decreases as with increasing particle size.

4.4 Bed fluctuation ratio

Effect of various parameters on bed fluctuation ratio:-

- With increasing bed height, the bed fluctuation ratio tends to increase.
- Larger particles tend to expand as a whole at higher liquid velocities.
- Smaller particles attain higher fluctuation ratio faster than the larger particles.
- Bed fluctuation ratio increases with gas velocity.
- Bed fluctuation ratio decreases with larger particles.

4.5 Future scope of the work

- To study the phase holdups of the three phase fluidized bed with irregular particles.
- Comparison of the bed behaviour of regular particles with irregular particles.
- To study the heat and mass transfer phenomena with irregular particles.

REFERENCES

- Chidambaram A. (2011). *CFD Analysis of Phase Holdup Behaviour in a Three Phase Fluidized Bed* (B.Tech. Thesis, National Institute of Technology, Rourkela, India).

- Jena, H.M., Sahoo, B.K., Roy, G.K., & Meikap, B.C. (2009 a). Statistical Analysis of the Phase Holdup Characteristics of a gas-liquid-solid Fluidized Bed. *Canadian Journal of Chemical Engineering*, 87(1), 1-10.

- Jena, H.M. (2009 b). *Hydrodynamics of Gas-Liquid-Solid Fluidized and Semi-Fluidized Beds* (PhD. Thesis, National Institute of Technology, Rourkela, India).

- Jena, H.M., Roy, G.K., & Meikap B.C. (2009 c). Hydrodynamics of a gas–liquid–solid fluidized bed with hollow cylindrical particles. *Chemical Engineering and Processing*, 48, 279-287.

- Jena, H.M., Sahoo, B.K., Roy, G.K., & Meikap B.C. (2008). Characterization of hydrodynamic properties of a gas–liquid–solid three-phase fluidized bed with regular shape spherical glass bead particles. *Chemical Engineering Journal*, 145, 50-56.

- Kim, S.P., Baker, C.G.J., & Bergougnou, M.A. (1972). Hold-Up and Axial Mixing Characteristics of Two and Three-Phase Fluidized Beds. *Canadian Journal of Chemical Engineering*, 50, 695-701.

- Levenspiel, O., & Kunii, D. (1991). *Fluidization Engineering* (2nd ed.), Boston: Butterworth-Heinemann. ISBN: 0-409-90233-0.

- Pandey, S.K. (2010). *CFD Simulation of Hydrodynamics of Three Phase Fluidized Bed* (M.Tech. Thesis, National Institute of Technology, Rourkela, India).

- Singh R.K., Suryanarayana A., & Roy G.K. (1999). Prediction of Bed Expansion Ratio for Gas-solid Fluidization in Cylindrical and Non-cylindrical Beds. *IE(1) Journal-CH*, 79, 51-54.

- Singh R.K., & Roy, G.K. (2006). Prediction of bed fluctuation ratio for gas solid fluidization in cylindrical and non-cylindrical beds. *Indian Journal of Chemical Technology*, 13, 139-143.